PB81-231243

Demonstration of Zinc Cyanide Recovery Using Reverse Osmosis and Evaporation

American Electroplaters' Society, Inc.
Winter Park, FL

Prepared for

Industrial Environmental Research Lab.
Cincinnati, OH

Jul 81

U.S. Department of Commerce
National Technical Information Service
NTIS

EPA-600/2-81-132
July 1981

PB81-231243

DEMONSTRATION OF ZINC CYANIDE RECOVERY USING REVERSE OSMOSIS AND EVAPORATION

by

Kenneth J. McNulty
John W. Kubarewicz
Walden Division of Abcor, Inc.
Wilmington, Massachusetts 01887
for
The American Electroplaters' Society, Inc.
Winter Park, Florida 32789

Grant No. R805300

Project Officer

Mary K. Stinson
Industrial Pollution Control Division
Industrial Environmental Research Laboratory
Edison, New Jersey 08817

INDUSTRIAL ENVIRONMENTAL RESEARCH LABORATORY
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO 45268

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA 22161

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO.	2.	3. RECIPIENT'S ACCESSION NO.
EPA-600/2-81-132	ORD Report	PB81 231243

4. TITLE AND SUBTITLE	5. REPORT DATE
Demonstration of Zinc Cyanide Recovery Using Reverse Osmosis and Evaporation	July 1981
	6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Kenneth J. McNulty, John W. Kubarewicz Walden Division of Abcor, Inc., Wilmington, MA 01887	8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS The American Electroplaters' Society 1201 Louisiana Avenue Winter Park, Florida 32789	10. PROGRAM ELEMENT NO. 1BB610
	11. CONTRACT/GRANT NO. R805300
12. SPONSORING AGENCY NAME AND ADDRESS Industrial Environmental Research Laboratory Office of Research and Development U.S. Environmental Protection Agency Cincinnati, Ohio 45268	13. TYPE OF REPORT AND PERIOD COVERED Final
	14. SPONSORING AGENCY CODE EPA/600/12

15. SUPPLEMENTARY NOTES

Project Officer: Mary Stinson (201) 321-6683

16. ABSTRACT

A field test was conducted to demonstrate closed-loop recovery of zinc cyanide at a job plating facility. Since the zinc cyanide bath operates at room temperature with very little evaporation from the bath, reverse osmosis (RO) treatment of the rinsewater must be supplemented by evaporation in order to achieve the volume reduction necessary for return of a concentrate to the plating bath. The permeate from the RO unit was recycled to the first rinse after plating while the distillate from the evaporator was recycled to the second rinse after plating. Continuous, unattended operation of this system was demonstrated with no adverse effects on plating quality.

The economics of the combined RO evaporation system were assessed for a system designed to provide rinsing equivalent to the present two-stage counter-current rinse at the demonstration site. The analysis showed that the total operating cost (including amortization) was somewhat less for the combined RO evaporation system than the evaporation alone. The minimum cost occurred for 90% water recovery in the RO system. However, credits for rinsewater recovery were insufficient to completely off-set the total operating cost of the recovery system.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Electroplating Waste Treatment Reverse Osmosis Evaporation	Zinc Cyanide Spiral-wound, PA-300 membrane	13B

18. DISTRIBUTION STATEMENT	19. SECURITY CLASS *(This Report)*	21. NO. OF PAGES
Release to public	Unclassified	
	20. SECURITY CLASS *(This page)*	22. PRICE
	Unclassified	

EPA Form 2220-1 (9-73)

DISCLAIMER

This report has been reviewed by the Industrial Environmental Research Laboratory - Cincinnati, U.S. Environmental Protection Agency, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the U.S. Environmental Protection Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory - Cincinnati (IERL) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report is a product of the above efforts. It was undertaken to demonstrate the effectiveness and economic feasibility of using reverse osmosis for closed-loop control of metal finishing rinse wastes under actual plant conditions. The reverse osmosis system concentrates the chemicals for return to the processing bath while purifying the wastewater for reuse in the rinsing operation. The results of the report are of value to R&D programs concerned with the treatment of wastewaters from various metal finishing, non-ferrous metal, steel, inorganic and other industries. Further information concerning the subject can be obtained by contacting the Nonferrous Metals and Minerals Branch of the Energy Pollution Control Division.

David G. Stephan
Director
Industrial Environmental Research Laboratory
Cincinnati

ABSTRACT

A field test was conducted to demonstrate closed-loop recovery of zinc cyanide rinsewater at a job shop plating facility. Since the zinc cyanide bath operates at room temperature with very little evaporation from the bath, reverse osmosis (RO) treatment of the rinsewater must be supplemented by evaporation in order to achieve the volume reduction necessary for return of a concentrate to the plating bath. The permeate from the RO unit was recycled to the first rinse after plating while the distillate from the evaporator was recycled to the second rinse after plating. Continuous, unattended operation of this system was demonstrated with no adverse effects on plating quality.

Spiral-wound PA-300 membrane modules were used in the RO unit. Periodic tests were conducted throughout the demonstration to characterize membrane performance under standard conditions. These tests indicated a gradual loss in membrane flux and rejection. After 3,000 hours of exposure to the rinsewater, the membranes were cleaned by flushing with a cleaning solution. The cleaning resulted in nearly complete restoration of flux and rejection. The gradual loss in membrane performance is thus attributable to fouling of the membrane by particulates in the rinsewater. Such fouling can be reduced by better pre-filtration and reversed by periodic cleaning.

The economics of the combined RO evaporation system were assessed for a system designed to provide rinsing equivalent to the present two-stage counter-current rinse at the demonstration site. The analysis showed that the total operating cost (including amortization) was somewhat less for the combined RO evaporation system than for evaporation alone. The minimum cost occurred for 90% water recovery in the RO system. However, credits for rinsewater recovery were insufficient to completely off-set the total operating cost of the recovery system.

This report was submitted in fulfillment of EPA Grant Number R805300 by the American Electroplaters' Society, Inc., (AES) under the partial sponsorship of the U.S. Environmental Protection Agency. This report covers the period August 24, 1977, to November 24, 1978, and work was completed as of February 7, 1979.

BEGIN FIRST LINE OF TEXT HERE
TYPING GUIDE SHEET
CENTER OF PAGE
TOP OF IMAGE AREA
DROPPED HEAD: BEGIN SECTIONS HERE

CONTENTS

Foreword iii
Abstract iv
Figures vi
Tables vii
Acknowledgment viii

1. Introduction 1
2. Conclusions 3
3. Recommendations 5
4. Methods and Materials 6
5. Results and Discussion 12

References 33

9-1/8"
BEGIN LAST LINE OF TEXT
BOTTOM OF IMAGE AREA; OUTSIDE DIMENSION FOR TABLES AND ILLUSTRATIONS
3/8"

PAGE NUMBER
EPA-287 (Cin.)
(4-76)

BEGIN FIRST LINE OF TEXT HERE
TYPING GUIDE SHEET
CENTER OF PAGE
TOP OF IMAGE AREA
DROPPED HEAD BEGIN SECTIONS HERE

FIGURES

Number		Page
1	Overall Schematic of RO/Evaporator Operation	7
2	Flow Schematic for RO Demonstration System	8
3	Flow Schematic for Evaporator	10
4	Flux as a Function of Exposure Time for Total Recycle Tests at 10% of Bath and 75°F	16
5	Zinc Rejection vs. Exposure Time as Determined by Total Recycle Tests at 10% of Bath	17
6	Free Cyanide Rejection vs. Exposure Time as Determined by Total Recycle Tests at 10% of Bath	19
7	Total Solids Rejection vs. Exposure Time as Determined by Total Recycle Tests at 10% of Bath	20
8	Dependence of Flux on Feed Concentration	23
9	Dependence of Rejection on Feed Concentration	24
10	Flows and Concentrations for Open and Closed-loop Rinsing	25
11	Annual Operating Costs for Various RO Conversions	32

BEGIN LAST LINE OF TEXT
3/8"
BOTTOM OF IMAGE AREA, OUTSIDE DIMENSION FOR TABLES AND ILLUSTRATIONS

PAGE NUMBER
EPA-287 (Cin.)
(4-76)

BEGIN FIRST LINE OF TEXT HERE
TYPING GUIDE SHEET
CENTER OF PAGE
TOP OF IMAGE AREA
DROPPED HEAD; BEGIN SECTIONS HERE

TABLES

Number		Page
1	Rejection Results for Evaporator During Closed-Loop Operation	14
2	Capital Costs for Various RO System Conversions	28
3	Total and Net Operating Cost for 90% RO System Conversion (5,000 Operating Hours Per Year)	29

6-1/2"
9-1/8"
BEGIN LAST LINE OF TEXT
BOTTOM OF IMAGE AREA, OUTSIDE DIMENSION FOR TABLES AND ILLUSTRATIONS
3/8"

PAGE NUMBER
EPA-267 (Cin.)
(4-76)

ACKNOWLEDGMENT

The authors acknowledge the cooperation and support of Bruce Warner, president of New England Plating Co. in providing the field site for the demonstration, support personnel to assist in installation, and cost information for operation and waste treatment at New England Plating Co. Technical support was received throughout the program from the EPA Project Officer, Ms. Mary Stinson, and from the AES Project Committee: Messrs. Jack Hyner, Joseph Conoby, Charles Levy, Herbert Rondeau, James Morse, and George Scott.

Financial support was provided both by EPA (Grant No. R805300) and by AES (Research Project No. 45).

SECTION 1

INTRODUCTION

Wastewater treatment technologies for the electroplating industry can be broadly classified as end-of-pipe destruction processes or in-plant recovery processes. The end-of-pipe destruction processes treat a total shop effluent to remove a mixture of heavy metals. At present it is neither technically nor economically feasible to recover and recycle metals from the end-of-pipe processes (1). On the other hand, in-plant recovery processes treat rinsewater from a specific plating bath (or other operation) making it possible to recover and return the heavy metals to the plating bath.

Because of the inherent disadvantage of end-of-pipe treatment -- loss of valuable plating chemicals, cost of treatment chemicals, cost of sludge disposal -- increasing attention has been focused on closed-loop recovery methods. In many cases, the economics of closed-loop recovery have been very favorable resulting in rapid payback on the capital investment for recovery equipment (2).

Aside from a few applications in which closed-loop recovery can be achieved by countercurrent rinsing alone, some technique must be used to remove the dissolved plating chemicals from the rinsewater. Although other techniques are under development, evaporation, reverse osmosis (RO), and ion exchange are the most commonly used processes for rinsewater recovery (1,3). Each of these techniques has particular advantages and disadvantages, and the best technique or combination of techniques will depend on factors specific to each application.

A number of advantages can be cited for the use of RO in rinsewater recovery. These include low capital cost, low energy and operating costs, and minimal space requirements. However, there are also some limitations. The major limitations for RO are:

1. The membrane modules deteriorate with time and require periodic replacement. The rate of deterioration depends on the type of membrane, the rinsewater pH and temperature, and the concentration of other reactants in the rinsewater such as oxidants.

2. Reverse osmosis cannot produce a highly concentrated stream for recycle to the plating bath. Thus for ambient temperature baths, RO must be supplemented with some other concentration technique, such as evaporation, in order to close the loop.

To date, RO has been applied primarily to the recovery of nickel rinsewaters. For nickel, the rinsewaters are relatively mild in pH (4-6) resulting in acceptable life for the conventional commercial membranes (cellulose acetate and aromatic polyamide). In addition, nickel baths operate at elevated temperatures where substantial evaporation occurs, and closed-loop operation can be achieved with RO alone.

Several programs, jointly sponsored by EPA and AES, have been conducted to evaluate the applicability of RO to plating baths other than nickel (4,5,6). Laboratory tests were conducted with a variety of newly developed membranes and rinsewaters with extreme pH levels (6). These tests indicated that of the membranes tested, the PA-300 was superior to the other membranes for treatment of copper cyanide, zinc cyanide, and chromic acid rinsewaters. The PA-300 membrane has since been commercialized (currently designated TFC-PA; manufactured by Fluid Systems Division of UOP) and is available in a spiral-wound modular configuration.

A field test was undertaken to evaluate the PA-300 membrane module for recovery of zinc cyanide rinsewater under realistic conditions. Zinc cyanide was selected because of the large volume of zinc cyanide plating done by the industry and because the high pH of the rinsewaters would provide a "worst case" test of the membrane for resistance to alkaline conditions. Since the zinc cyanide bath operates at room temperature, it was necessary to use an evaporator to supplement RO treatment and achieve the level of concentration necessary for closed-loop operation. This report presents and discusses the results of this field test.

SECTION 2

CONCLUSIONS

Closed-loop recovery of zinc cyanide rinsewaters can be achieved with a combined RO/evaporator system. Continuous and unattended operation of the system was demonstrated over one-week periods (Monday through Friday, three shifts per day). No adverse effects on plating quality were noted during the demonstration.

The single-effect vacuum evaporator was operated at about one-half of its rated capacity, and the vacuum was sufficient to keep the vaporization temperature below 43°C (110°F). The quality of the distillate was considered quite suitable for final rinsing.

Tests were conducted periodically to determine the flux and rejection performance of the three PA-300 spiral-wound modules used in the RO system. Results for Modules #1 and #2 were in close agreement throughout the demonstration. Module #3 was concluded to be defective from the outset of the program and was removed after 3,000 hours.

The flux and rejection of modules #1 and #2 gradually declined during the demonstration. Over the first 3,000 hours of exposure, the flux declined to about one-half of its original value and the zinc rejection, for example, declined from 99% to 97%. At 3,000 hours, the membranes were cleaned by flushing with a cleaning solution. The cleaning resulted in nearly complete restoration of flux and rejection. The gradual loss in membrane performance is thus attributable to fouling of the membrane by particulates in the rinsewater. Such fouling can be reduced by better prefiltration and reversed by periodic cleaning.

Following the field test, Module #2 was cut open and unwound for inspection. The membrane was fouled with a thin layer of what appeared to be sulfide sludge. (Polysulfide is used as a bath purifier). Examination of the module internals revealed that some deterioration in the strength of the membrane backing material had occurred. However, this produced no gross effects on the performance of the modules as observed during the field test.

The economics of the combined RO evaporation system were assessed for a system designed to provide rinsing equivalent to the present two-stage counter-current rinse at the demonstration site. The analysis showed that the total operating cost (including amortization) was somewhat less for the combined RO evaporation system than for evaporation alone. Total operating costs were calculated for various RO system water recoveries, and the minimum cost occurred for 90% water recovery in the RO system. However

credits for rinsewater recovery were insufficient to completely off-set the total operating cost of the recovery system. Energy costs for evaporation using a double effect evaporator were 4-5 times greater than those for RO. Continuing escalation of energy costs would provide further incentive for a combined RO evaporation system.

SECTION 3

RECOMMENDATIONS

On the basis of this field test and previous laboratory tests (6) the PA-300 membrane can be recommended for the treatment of cyanide rinsewaters on a commercial scale. Application to copper cyanide recovery appears to be particularly attractive because of: 1) the lower pH of copper cyanide relative to zinc cyanide, 2) the higher value of copper relative to zinc, and 3) the higher evaporation rate from the copper bath which would permit closed-loop operation with RO alone (no evaporator required).

Additional development of the PA-300 or similar membranes is recommended with particular emphasis on the development of modules containing materials which are highly resistent to pH extremes. In particular, a more alkaline-resistent membrane backing material should be developed for cyanide applications.

Data should be obtained on the life of PA-300 (or similar) modules for longer exposure times (>4,200 hours). This data can be conveniently and inexpensively obtained by "soaking" the module in a solution of the plating bath diluted to simulate concentrated rinsewater. The flux and rejection of the modules can then be determined periodically. These soak tests will provide information on the resistance of various module components to attack by the major constituents of the bath (e.g. OH^-).

In view of the good acid resistance exhibited by the PA-300 membrane in laboratory tests (6), it is recommended that a field test and long-term soak tests be conducted using PA-300 modules and low pH rinsewater such as acid copper. (In future field tests, the feed to the modules should be pretreated by passage through a 1 μ filter). The objective of these tests would be to extend the applicability of RO to pH levels below 2.5 (the lower limit for cellulose acetate membranes).

For evaporative recovery, development of a low capacity, low cost, mechanical vapor recompression evaporator is recommended. This type of evaporator is highly efficient and could be operated at a significantly lower cost than the combined RO/evaporator system tested in this program.

SECTION 4

METHODS AND MATERIALS

A mobile RO test system was leased from Abcor, Inc., and an evaporator was leased from Wastesaver Corporation for the duration of the field test. These two units were installed on an automatic rack, zinc cyanide plating line at New England Plating Co. in Worcester, Massachusetts. The overall schematic of the installation is shown in Figure 1. Feed to the RO system was withdrawn from Rinse Tank #1 and separated by the RO system into a permeate stream and a concentrate stream. For purposes of design, it was assumed that the RO system would produce about 7.5 lpm (2 gpm) of permeate and would operate at 90% conversion. (Conversion is defined as the ratio of permeate flow to feed flow.) Thus the RO system would be fed at the rate of 8.4 lpm (2.22 gpm) and would produce concentrate at the rate of .8 lpm (0.22 gpm). The permeate was returned to Rinse Tank #1 and the concentrate was fed to the evaporator.

Since dragin and dragout were essentially identical for the plating bath and the rate of evaporation was negligible, there was no room in the plating bath for a concentrate stream. If the evaporator were fed only RO concentrate, it would have to evaporate it to dryness in order to prevent eventual overflow of the bath. In order to prevent precipitation of plating chemicals in the evaporator a 3.8 lpm (1 gpm) purge stream was circulated from the plating bath through the evaporator and carried the plating salts introduced with the RO concentrate back to the plating bath. That is, the evaporator concentrate was higher in concentration than the plating bath by the amount added by the RO concentrate. The distillate from the evaporator was collected in a holding tank and added at a controlled rate to Rinse Tank #2. A float valve operating off the level in Rinse Tank #1 insured that the rate of RO concentrate production was exactly balanced by the rate of distillate returned to Rinse Tank #2. A slight excess of distillate was produced to insure that the holding tank would always remain full; and the excess was permitted to overflow into the plating bath (.08 lpm). The steam rate was cut back to minimize overflow from the holding tank.

A flow schematic of the RO system is shown in Figure 2. Feed from the first rinse tank was withdrawn by a booster pump and passed through two string-wound cartridge filters in parallel. Both 1-μ and 20-μ filters were used at different times during the field test. Excess flow from the booster pump was returned to the rinse tank. After pre-filtration, the feed was pressurized to 48.6 atm (700 psi) with a multi-stage centrifugal feed pump and passed through three 102 mm (4-inch) diameter, spiral-wound, PA-300 modules arranged in series. Most of the concentrate from the third module was recycled to the suction of the feed pump in order to maintain the required feed flow rate through the modules. A heat exchanger in the recir-

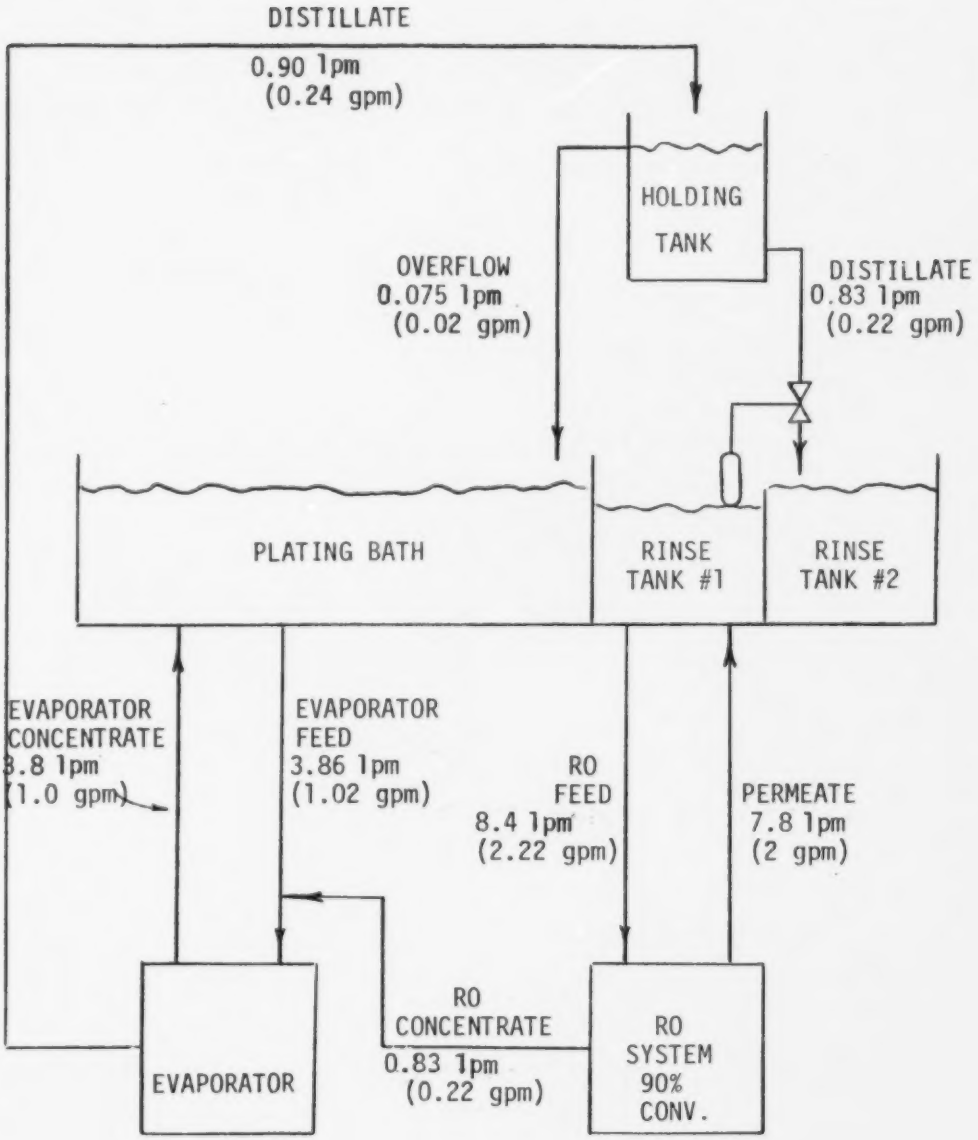


Figure 1. Overall Schematic of RO/Evaporator Operation.

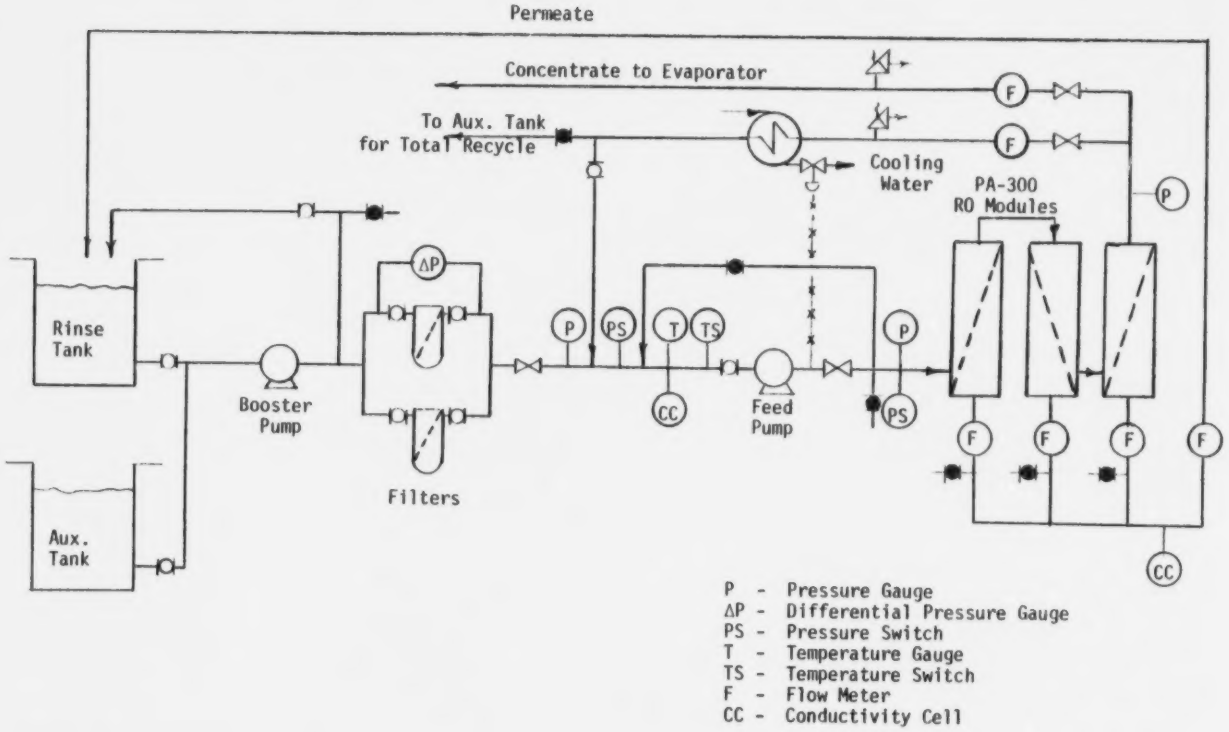


Figure 2. Flow Schematic for RO Demonstration System.

culation loop removed heat generated by the energy input of the pumps. A small flow of concentrate from the third module was fed to the evaporator (see Figure 1), and the permeate from the three modules was combined and returned to the first rinse tank. The instrumentation and controls for the RO system are shown in Figure 2.

In order to characterize membrane performance with a standard feed solution, the RO system was periodically operated in a total recycle mode using the auxiliary feed tank. For this mode of operation, the booster pump recycle line was closed off, the concentrate line to the evaporator was opened, and the permeate was returned to the auxiliary tank rather than the rinse tank. The standard solution (generally a portion of plating bath diluted to 10% by volume or original bath strength) was charged to the auxiliary tank and the system was operated with total recycle until steady state was achieved. At steady state, the permeate flow rate for each module was measured, and samples of the feed and permeate from each module were obtained for analysis.

Typical operating conditions for both closed-loop and total recycle were:

Feed Pressure	48.6 atm (700 psi)
Recirculation Flow Rate	37.8 lpm (10 gpm)
Temperature	21-32°C (70-90°F)
Concentrate Flow Rate	0.75 lpm (0.2 gpm) closed-loop only

The flow schematic for the evaporator is shown in Figure 3. Steam was fed through a pressure reducing valve to a tube bundle submerged in the boiler section of the evaporator, and steam condensate was returned to the plant boiler. For most installations, a cooling tower is used to cool the water which is recirculated through the condenser section of the evaporator. However, for this installation it was more convenient to use recirculated chilled water since it was readily available at the installation site and the chiller had sufficient excess capacity. The evaporator was maintained under vacuum by circulating water through an eductor. Cooling water was added to the eductor tank to remove the energy input of the eductor circulation pump. Feed to the evaporator was controlled by a level switch (LS) and solenoid valve. Upon low level signal, the solenoid valve opened and feed was drawn be vacuum into the evaporator. The distillate from evaporation of the feed condensed, was collected in a tray below the condenser, and was continuously pumped back to the second rinse after plating (see Figure 1). The concentrate from the boiler section of the evaporator was continuously pumped back to the plating bath.

Typical operating conditions for the evaporator were:

Vacuum	0.87 atm abs. (26-27 in. Hg)
Temperature	38-43°C (100-110°F)
Steam Pressure	<1.3 atm abs. (<19.7 psig)
Concentrate Flow Rate	3.8 lpm (1 gpm)

During the field test, the RO modules were cleaned using a cleaning sequence recommended by the membrane manufacturer. The modules were first flushed with 189 liters (50 gal) of water to remove the plating chemicals. A

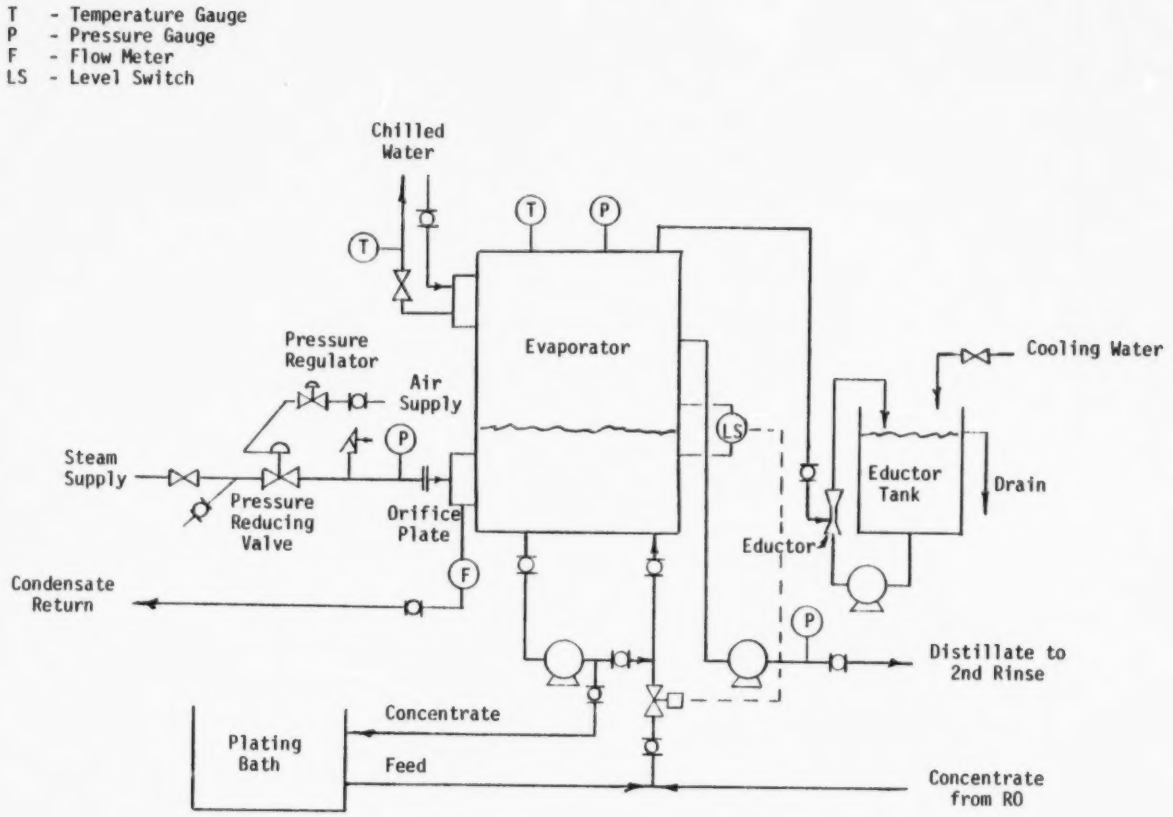


Figure 3. Flow Schematic for Evaporator.

0.1% by volume solution of Triton X-100, a non-ionic surfactant, was prepared and recirculated through the modules at a pressure of 48.6 atm abs (700 psi), a flow rate of 37.9 lpm (10 gpm), and a temperature of 49°C (120°F) for 45 minutes. After flushing with another 189 liters of water, a 2% citric acid solution was prepared and adjusted to pH 3.0 with ammonium hydroxide. This solution was recirculated through the modules at the same conditions and for the same time as the Triton X-100. Following the citric acid cleaning the system was again flushed with water and returned to treatment of zinc cyanide rinsewater. Since the PA-300 membrane is rapidly degraded by chlorine, all water used for flushing and preparing cleaning solutions was dechlorinated by the addition of sodium sulfite.

Samples collected during the field test were analyzed for zinc (atomic absorption), free cyanide (selective ion electrode), total solids (gravimetric determination of residue), conductivity (conductivity bridge), and pH (electrode).

The nominal composition of the plating bath was:

Zn (as metal)	20,000 mg/l	2.7 oz/gal
CN (as NaCN)	60,000 mg/l	8.0 oz/gal
Caustic	75,000 mg/l	10.0 oz/gal
Brightener (700 Special)	4 ml/l	4 gal/1000 gal

In addition to these compounds, polysulfide was regularly added to the bath for purification, and the bath also contained a large quantity of carbonates. The total solids concentration of the bath was in the vicinity of 350,000 mg/l (35% by weight).

SECTION 5

RESULTS AND DISCUSSION

Field activities covered the months of June through December of 1978. During this time, the system was operated primarily in the closed-loop mode shown in Figure 1. However, closed-loop operation was periodically interrupted to conduct total recycle tests on the RO system. The performance of RO modules is generally assessed in terms of the module flux (rate of permeate production per unit membrane area) given in liters per square meter per hour l/m^2hr or gallons per ft^2 per day (gfd) and the rejection (defined for a particular dissolved species as one minus the ratio of permeate to feed concentrations) given in percent. Since both flux and rejection depend on feed concentration, it is necessary to conduct RO performance tests at a fixed feed concentration in order to be able to accurately interpret trends in the flux and rejection data. During closed-loop operation the feed concentration to the RO system can vary considerably (depending on the parts being plated and their dragout); hence total recycle tests at a fixed feed concentration (10% of bath strength) were periodically conducted to accurately assess trends in the RO system performance. System performance was monitored less rigorously during operation in the closed-loop mode.

The total operating time for the field demonstration (time during which permeate was being produced) was approximately 1000 hours. The total exposure time of the modules to the rinsewater, however, was about 4,200 hours. The longer exposure time reflects the system down-time during which the modules were sitting in contact with the concentrated zinc cyanide rinsewater. Various factors contributed to the system down-time, including: electro-mechanical failure of various system components, high or low alarm shut-down of the system, various problems with the plating line (unrelated to the recovery system), and weekends and holidays. The total exposure time is probably more significant than operating time in controlling the degree to which the modules are attached chemically by constituents in the rinsewater (e.g., caustic). On the other hand, operating time is more significant in controlling the degree to which the modules become fouled with particulates in the feed. Correlations presented below are based on exposure time. However, the operating time was reasonably evenly spread over the test program, and correlations on an operating-time basis would be similar.

Closed-Loop Operation

The RO-evaporator system was designed to operate continuously, with no operator attention, between start-up Monday morning and shut-down Friday afternoon. However, during most of the field test program, various

electro-mechanical and other problems occurred which prevented unattended week-long operation. Each failure generally resulted in several days down time because of the logistics of getting project personnel to the field site, diagnosing the problem, and implementing remedial action. Eventually these problems were solved and week-long unattended operation was demonstrated.

During demonstration of the closed-loop system, no adverse effects were noted on the quality of the plated parts. However, the rinse tanks after zinc plating were followed by an acid dip, a flowing rinse, and other surface finishing steps before the parts were finally inspected. Therefore the degree of rinsing following zinc cyanide plating was probably not of critical importance to quality control.

Both the RO system and evaporator were underdesigned as a direct replacement for the two-stage countercurrent rinse at a nominal rinsewater flow rate of 7.8 lpm (2 gpm) (see "Economics" below). The system design reflected limitations imposed on program costs, availability of PA-300 modules at the time the system was fabricated, and lack of design data for the zinc cyanide application. Nevertheless, the system was of sufficient size to obtain meaningful design and economic data.

RO performance during closed-loop operation was monitored by measuring the productivity of each module (rate of permeate production) and by monitoring the conductivity of the combined RO permeate. In general, the productivity during closed-loop operation was similar (but slightly higher) than the productivity during the total recycle tests (see discussion below). The conductivity of the combined permeate generally ranged between 2,000 and 4,500 μ-mhos/cm. This is equivalent to a total solids concentration of approximately 1,000 to 2,000 mg/l.

The evaporator was operated at about one-half of its rated capacity (57 liters per hour vs. a capacity of 95 lph). Samples of distillate and evaporator concentrate were obtained and analyzed for zinc, free cyanide, total solids, conductivity and pH. The results of these analyses are presented in Table 1. During closed-loop operation at the time the samples were taken, a stream of about 5.7 lpm (1.5 gpm) from the bath was circulated through the evaporator and back to the bath in order to prevent precipitation of plating salts in the evaporator. Thus, the evaporator was operating on a feed very similar in composition to the bath and producing a concentrate which was more concentrated than the plating bath. (In addition, the plating bath during this test appears to be significantly higher in zinc than the nominal level given above). The quality of distillate produced was quite good. Zinc and total solids concentrations were below their respective detection limits, but cyanide, conductivity, and hydroxide ion were detectable. The rejections or removal efficiencies were quite high; particularly for zinc and total solids. The quality of distillate was considered quite suitable for final rinsing.

TABLE 1. REJECTION RESULTS FOR EVAPORATOR DURING CLOSED-LOOP OPERATION

	CONCENTRATE	DISTILLATE	REJECTION
Zinc	46,000 mg/l	<0.1 mg/l	>99.9998%
Cyanide	90,000 mg/l	35 mg/l	99.96%
Total Solids	409,000 mg/l	<5 mg/l	>99.999%
Conductivity	160,000 mg/l	300 mg/l	99.81%
pH	13.4	10.3	

NOTE: Operating conditions 40°C (104°F), .9 Atm abs (27" Hg vacuum), 17% Conversion.

Total Recycle Tests at 10% of Bath Concentration

Total recycle tests were periodically conducted in order to evaluate membrane module performance under well defined conditions of feed concentration (10% of bath concentration), pressure (48.6 atm abs., 700 psig), temperature (24-29°C, 75-85°F), and recirculation rate through the modules (37.8 lpm, 10 gpm). Results for flux as a function of time are shown in Figure 4. The flux was calculated from the measured permeate flow rate using a surface area of 6.5 m^2 (70 ft^2) per module. (Actual surface areas measured after the field test for Modules #2 and #3 were 6.8 m^2 (73 ft^2) and 6.4 m^2 (69 ft^2), respectively.) the data were corrected to 24°C (75°F) using the inverse relation between flux and water viscosity.

For Modules #1 and #2 the flux levels were nearly identical throughout the field demonstration. The flux gradually declined from 22 to 12 l/m^2hr (13 gfd to 7 gfd) over the first 3,000 hours. Following cleaning at 3,000 hours, the flux increased to 20 l/m^2hr (12 gfd), which is very close to the original level. Thus the cleaning procedure employed was quite successful in restoring the flux. It may therefore be concluded that the major reason for flux decline is fouling of the membranes.

Results for Module #3 are also presented in Figure 4 (dashed line). The flux starts at a significantly higher level, 34 l/m^2hr (20 gfd) but declines more rapidly to a value of 8.5 l/m^2hr (5 gfd) at 3,000 hours. The cleaning procedure was ineffective in restoring the flux of Module #3. Because of its low flux and low rejections (discussed below) Module #3 was removed from the system at 3,000 hours.

After cleaning, the system was returned to closed-loop operation on the actual rinsewater. During the next 1,200 hours the flux for Modules #1 and #2 declined to about 12 l/m^2hr (7 gfd). The rate of flux decline was more rapid during this time period since a coarser grade (20μ vs 1μ) cartridge filter was used to pre-filter the feed during this segment of the demonstration.

Results for zinc rejection during the field demonstration are shown in Figure 5. For Modules #1 and #2 the zinc rejections agree reasonably well and are correlated with a single curve (solid line). The zinc rejection declined gradually from an initial value of 99% to 97% after 3,000 hours. Upon cleaning, the rejection for Module #1 increased to nearly 99% while the rejection for Module #2 increased to about 98%. Thus the cleaning procedure was successful in restoring the zinc rejection. Therefore it is reasonable to conclude that the loss of zinc rejection is primarily the result of membrane fouling. After cleaning the rejection again declined as fouling occurred.

For Module #3 the zinc rejection dropped off rapidly to a value of 90% after only 700 hours and remained reasonably constant between 700 and 3,000 hours. Cleaning produced a significant increase in rejection (from 90 to 95%).

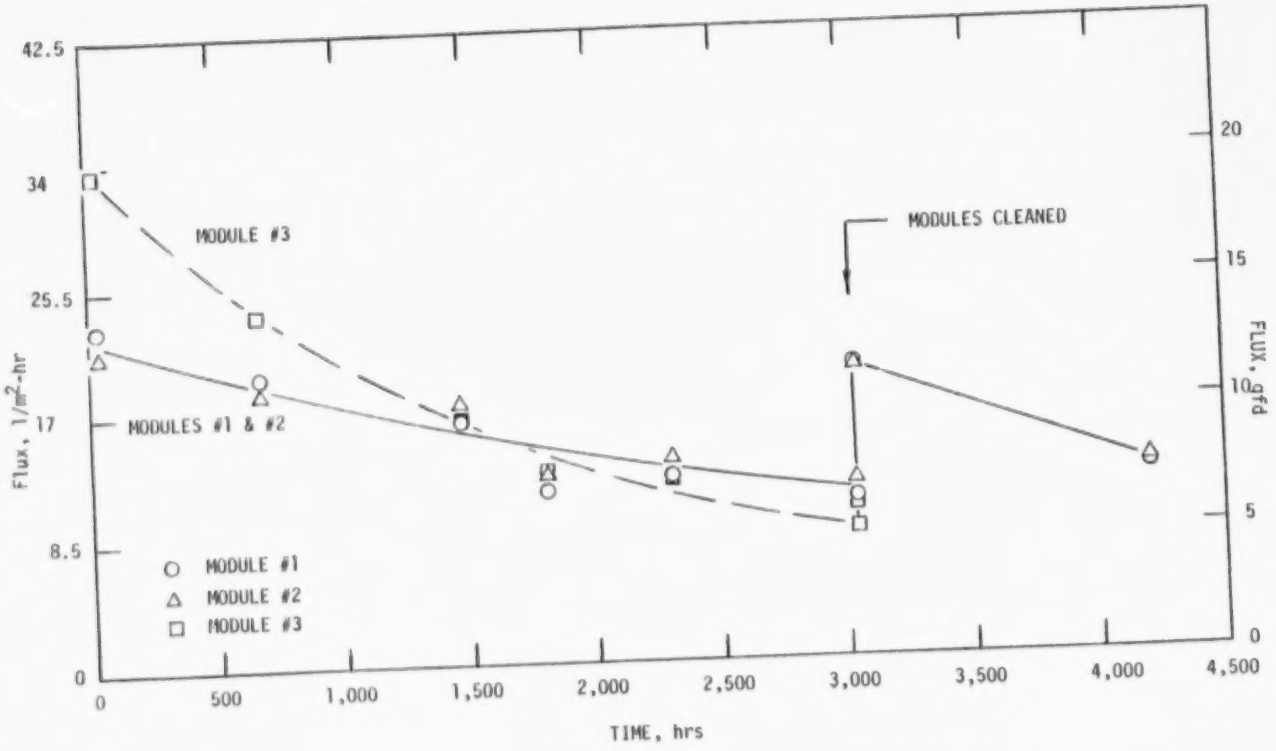


Figure 4. Flux as a Function of Exposure Time for Total Recycle Tests at 10% of Bath and 75° F.

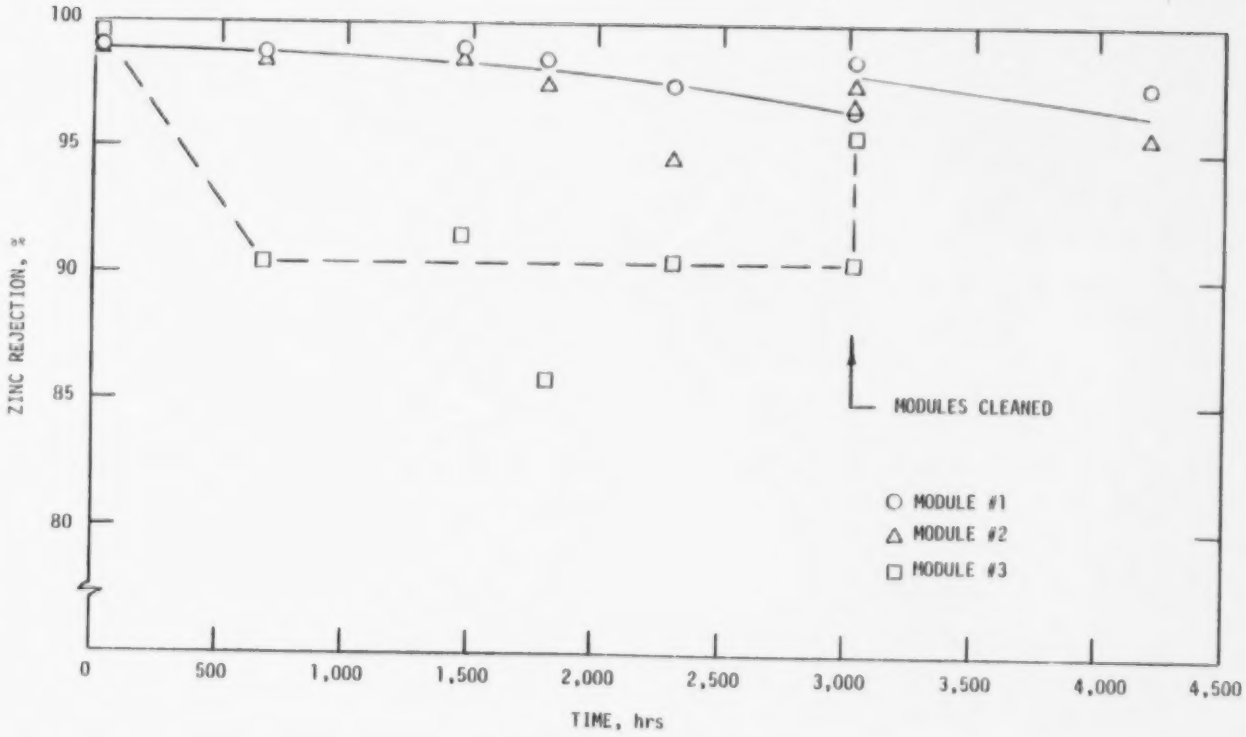


Figure 5. Zinc Rejection vs. Exposure Time as Determined by Total Recycle Tests at 10% of Bath.

The rejection of free cyanide is shown in Figure 6. Again the results for Modules #1 and #2 agree quite well throughout the test. For these modules the cyanide rejection declined slightly -- from 98% to 97% -- during the first 3,000 hours. Cleaning had little effect on the rejection level. However, because of the small loss in rejection, cleaning would be anticipated to have only a minor effect on rejection levels. Following cleaning, the cyanide rejection declined at a more rapid rate, probably as the result of the increased rate of fouling.

The cyanide rejection results for Module #3 are also shown in Figure 6. As for zinc, the cyanide rejection dropped quite rapidly, from 99 to 85%, during the first 700 hours and appeared to stabilize at about 85% until 3,000 hours. The data point at 3,000 hours (93% rejection) indicates a substantial recovery in rejection before cleaning. Since no similar recovery was observed for zinc and total solids rejections, it is possible that the high cyanide rejections at 3,000 hours could be attributed to a sampling or analytical error.

Total solids rejections are presented in Figure 7. The results for Modules #1 and #2 are, as before, almost identical. There was a gradual loss in total solids rejection for Modules #1 and #2 from 95% initially to 90% after 3,000 hours. Cleaning at 3,000 hours produced an apparent loss in rejection. (A similar loss was also noted for conductivity rejection.) During operation after cleaning, the rejection increased to 90% -- the same level as before cleaning. Thus the loss of rejection upon cleaning was only temporary. Indeed, since the results of Figures 4-6 indicate fouling and a decline in membrane performance between 3,000 and 4,200 hours, it is reasonable to postulate that a similar decline in total solids rejection occurred during this period. By virtue of the fact that the rejection at 4,200 hours is the same as at 3,000 hours, it can be concluded that the cleaning actually improved the total solids rejection. The observed loss in rejection is probably the result of an interaction between the citric acid cleaning solution and the membrane surface. Similar results have been observed with citric acid in cleaning tests conducted by the manufacturer (7). These tests consistently showed a loss in rejection after cleaning, but the rejection then increased over a relatively short period (5-24 hours) to the level expected for a clean module. The mechanism of this interaction is not well understood, but the rejection loss appears totally reversible by extensive flushing or by returning the system to operation on the normal feed. Thus the most reasonable explanation of the total solids rejection behavior is that, when the system was returned to normal operation after cleaning, the rejection increased to a level probably close to 95% within 24 hours and then decreased gradually to 90% at 4,200 hours as fouling of the module occurred.

The total solids rejections for Module #3 are quite similar to the zinc and cyanide rejections for this module. The rejection decreased from 95% to about 80% during the first 700 hours and remained at about 80% until the module was removed at 3,000 hours. Cleaning produced no significant change in total solids rejection for Module #3.

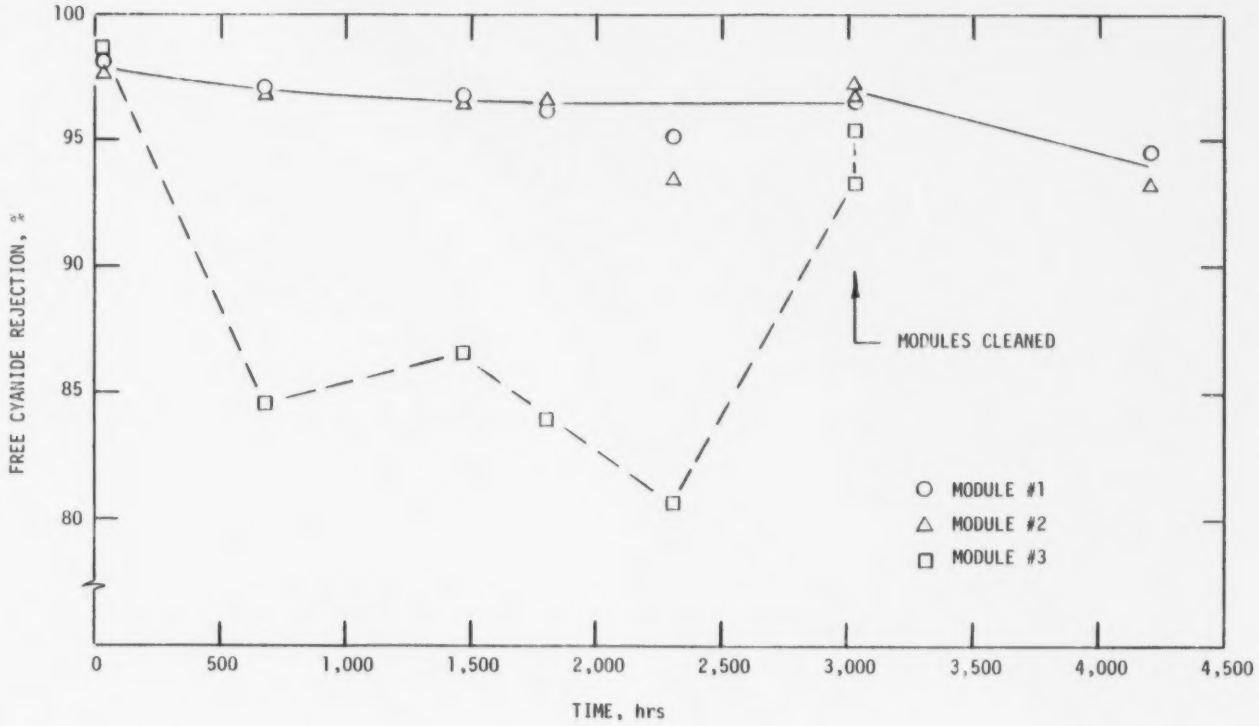


Figure 6. Free Cyanide Rejection vs. Exposure Time as Determined by Total Recycle Tests at 10% of Bath.

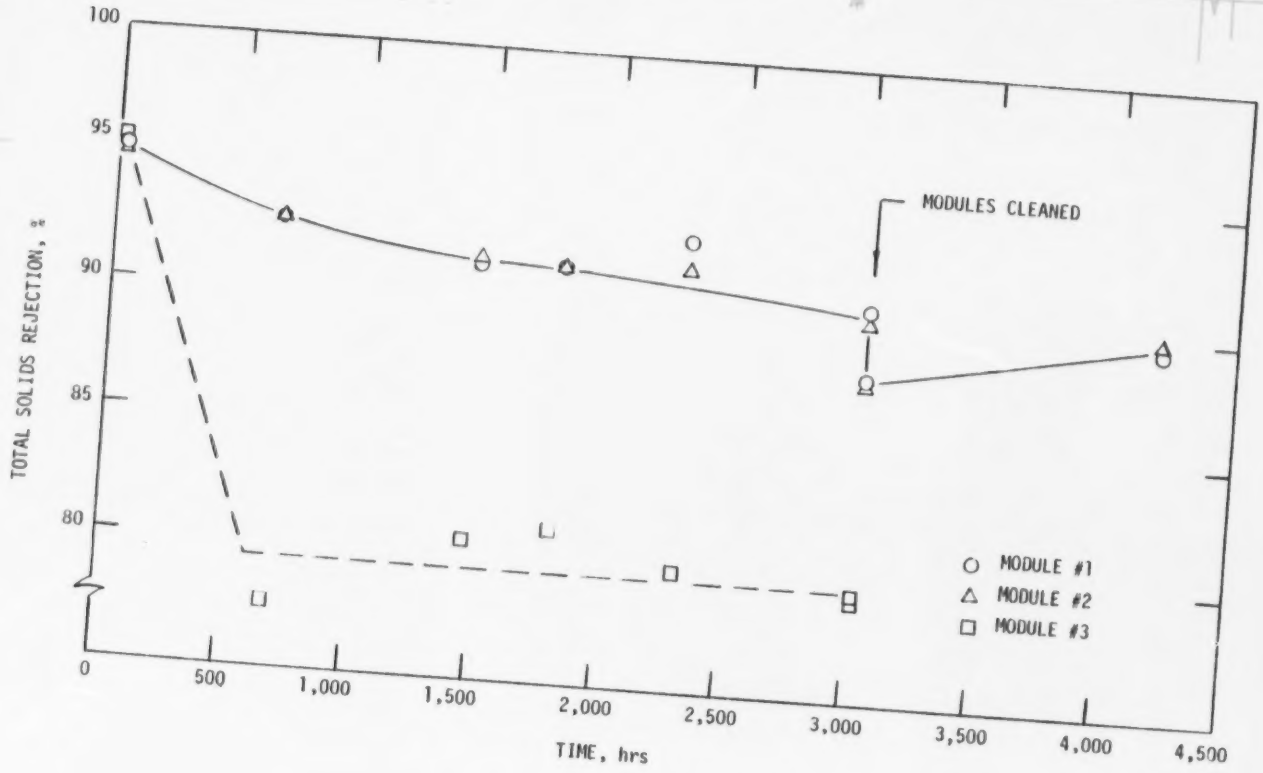


Figure 7. Total Solids Rejection vs. Exposure Time as Determined by Total Recycle Tests at 10% of Bath.

The close agreement between the results for Modules #1 and #2 would be anticipated for two identical modules operated in series. By comparison the results for Module #3 are quite poor. Since the system conversion per pass was low (high recirculation flow relative to permeate flow) the feed concentrations to the three modules were approximately the same; therefore a higher feed concentration would not account for the poor performance of Module #3.

A similar rapid loss in performance has been occasionally noted for some of the earlier PA-300 modules used in water desalination (7). This problem was traced to a procedure used in manufacturing the modules. The procedure has since been changed and the problem thereby eliminated. However, the modules used in this field test were manufactured before this change, and it is believed that Module #3 was defective from the outset of the test.

The level of suspended solids in the zinc cyanide bath was high compared to the levels observed for copper cyanide baths (5). A purifier (polysulfide) was regularly added to the zinc cyanide bath and produced a mud-like sludge that was removed by filtration. Some of this sludge was carried over into the rinse tanks and thus into the RO system. During the first 3,000 hours of the test, 1μ cartridge filters were used and had to be changed approximately once per week (assuming 100 operating hours per week). After 3,000 hours, 20μ cartridge filters were used. The service time for the coarser filters was substantially longer (estimated service time = one month) but the rate of fouling of the modules was greater with the coarser filters. It is recommended that two cartridge filters in series -- a 20μ filter followed by a 1μ filter -- be used for pre-filtration of zinc cyanide rinsewater.

Following the field test Module #2 was cut open and unwound for inspection. The membrane was fouled with a thin layer of sulfide sludge with the degree of fouling increasing toward the center of the module. This layer could be very easily wiped from the membrane surface. During cleaning at 3,000 hours, a distinct odor of H_2S was noted during recycling of the citric acid solution. This is presumed to be the result of solubilization of some of the sulfide foulant. This suggests that the ammoniated citric acid was the active ingredient in the cleaning procedure and that the Titron X-100 surfactant could be eliminated. Cleaning agents other than citric acid could work as well or better and should be tested. Of particular interest would be the use of oxidizing agents to oxidize the sulfide foulant layer. While the membrane is succeptible to rapid attack by some oxidizing agents such as chlorine, it has been shown to be resistant to others such as chlorine dioxide and chromic acid (6,7).

In addition to fouling of the membrane, examination of the module internals revealed some possible deterioration of the membrane backing material. Samples of the membrane and backing were returned to the manufacturer for examination, the finding of which confirmed that the backing had been deteriorated, probably as a result of the high concentrations of hydroxide ion in the rinsewaters (7). However, it is not clear what effect deterioration of the backing would have on membrane

module performance. No gross effects on performance were observed during the field test. Substitute backing materials could be used, but a development program would be required to commercialize the PA-300 on a more resistant backing.

Total Recycle Tests at Other Feed Concentrations

Following initial start up of the RO system, total recycle tests were conducted to determine the dependence of flux and rejection on feed concentration. Tests were conducted at various dilutions of the plating bath ranging in nominal concentration from zero to 20% of the bath concentration.

Flux results as a function of the total solids concentration of the feed are shown for Modules #1 and #2 in Figure 8. The results for these two modules are in very good agreement. (Results for Module #3 were not consistent with those shown in Figure 8 and have not been included). As expected, the flux decreases with increasing feed concentration and approaches a level generally considered uneconomical at a feed concentration in the vicinity of 100,000 mg/l. By comparison the bath concentration is approximately 350,000 mg/l in total solids. These data illustrate the problem of using RO to achieve very high concentrations. The data presented in Figure 8 can be used in optimizing the degree of concentration that should be obtained in the RO system prior to evaporation.

Rejection as a function of the total solids concentration in the feed is shown in Figure 9 for Modules #1 and #2. The zinc, cyanide, and total solid rejections remain essentially constant at 99%, 98%, and 95%, respectively for feed concentrations below about 40,000 mg/l total solids. At higher feed concentrations the rejections drop off as expected.

Economics

One of the main objectives of conducting a field demonstration is to provide a basis for evaluating the economics of the process. Figure 10a shows the flow schematic for the original two-stage countercurrent rinsing system. The rinsewater flow rate was measured at 7.6 lpm (2.0 gpm) The rate of dragout from the bath was estimated by turning off the rinsewater and measuring the buildup of zinc and total solids in the first rinse as a function of time. The calculated dragout rate was .04 to .06 lpm (0.01 to 0.015 gpm). However the parts being plated during this determination were flat and free-draining and a somewhat higher dragout rate of .08 lpm (0.02 gpm) was considered more typical for purposes of design. Given these flow rates and ideal mixing in the rinse tanks, the zinc concentration is reduced from 20,000 mg/l in the bath to 2.0 mg/l in the final rinse.

Figure 10b shows the flow schematic, flow rates (in lpm), and zinc concentrations (in mg/l) for an RO/Evaporator system designed to give the same zinc concentration in the final rinse. The flow rates given in Figure 10b are for 90% water recovery in the RO system (ratio of system permeate to system feed: $\frac{22.0}{24.6} = .90$). All of the concentrate from the

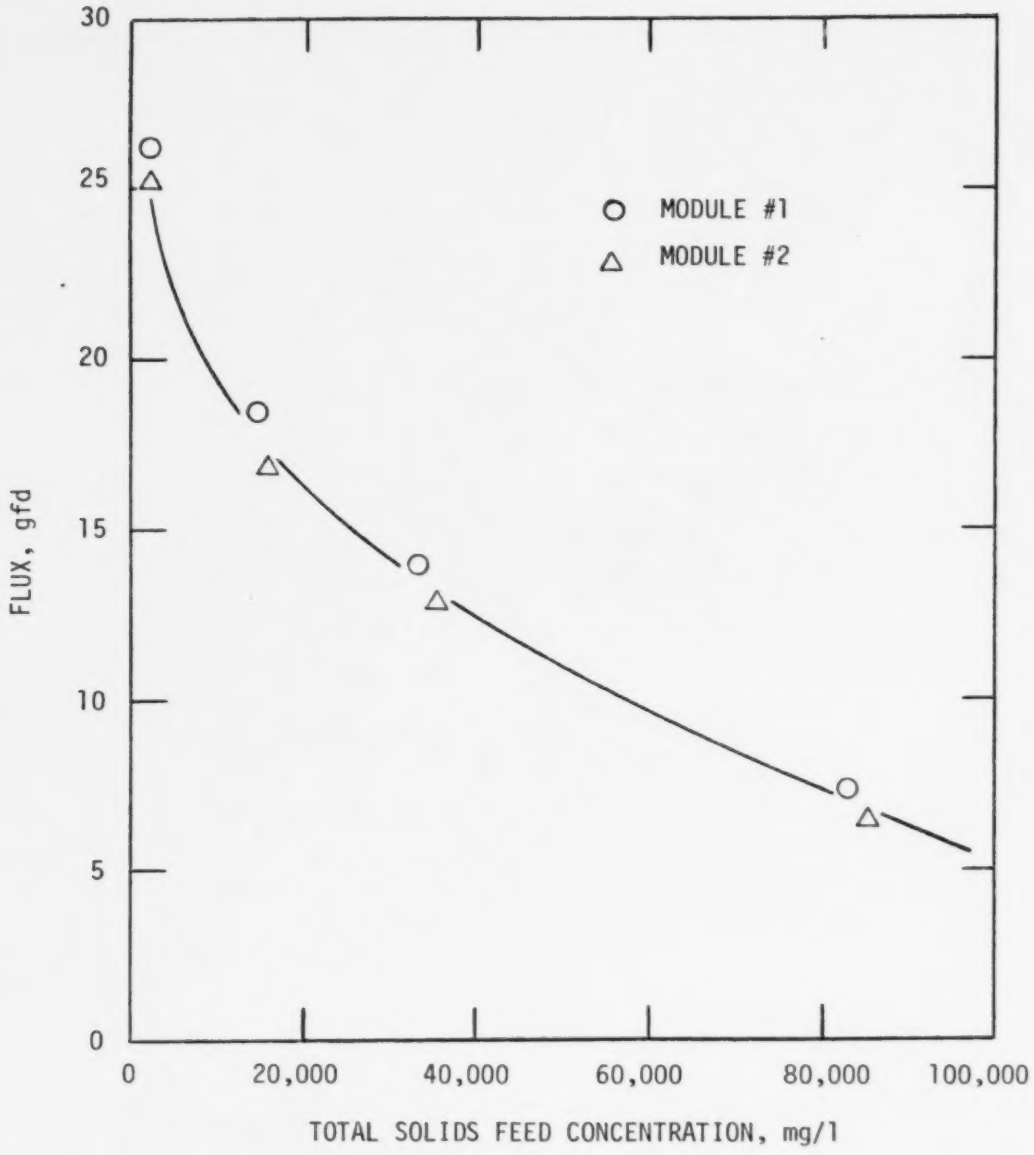


Figure 8. Dependence of flux on feed concentration.

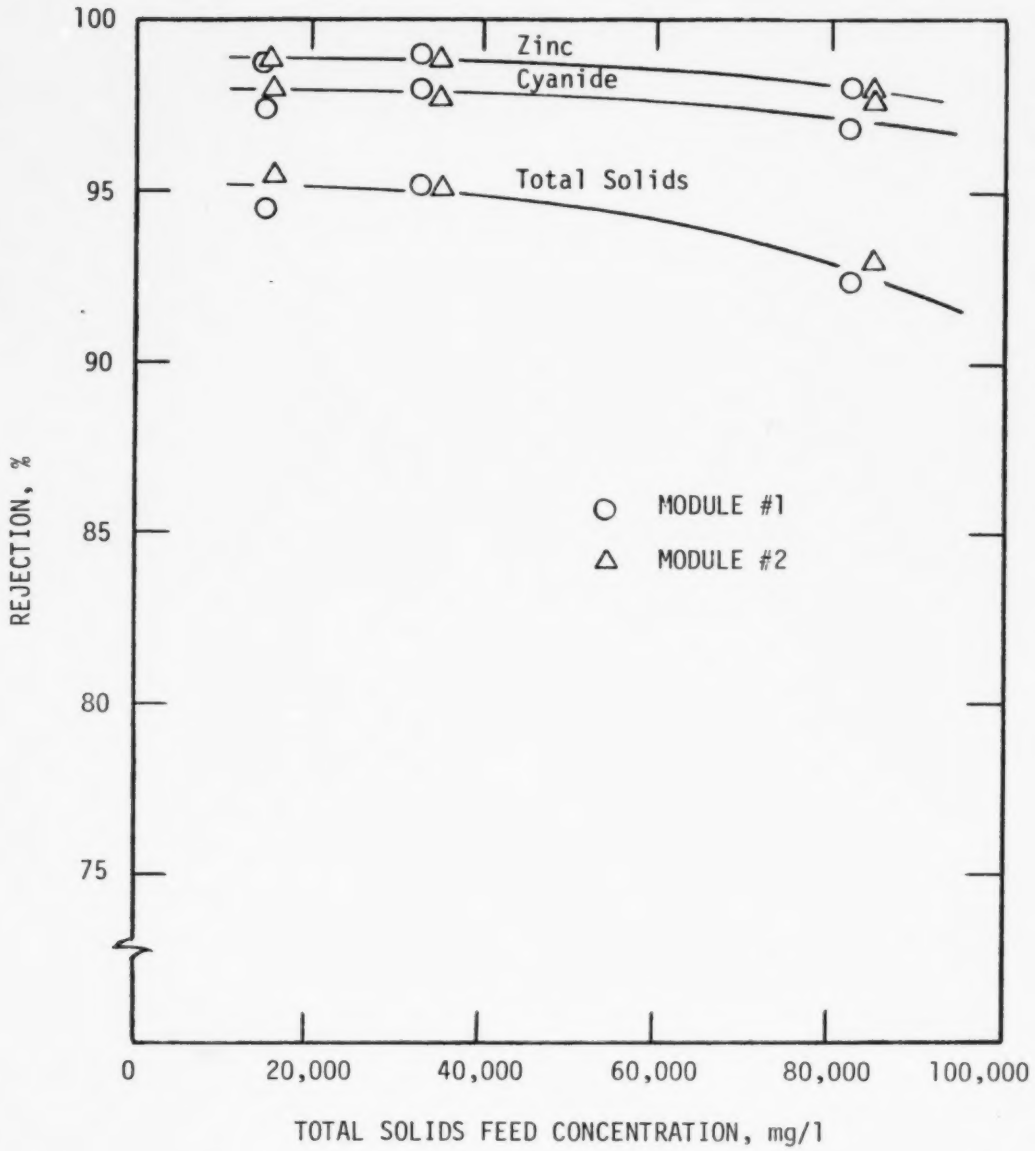
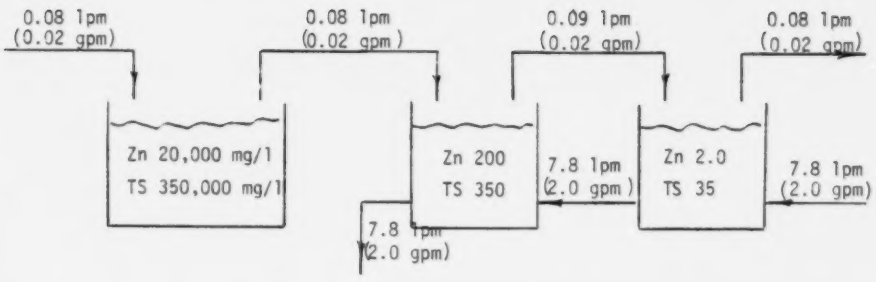
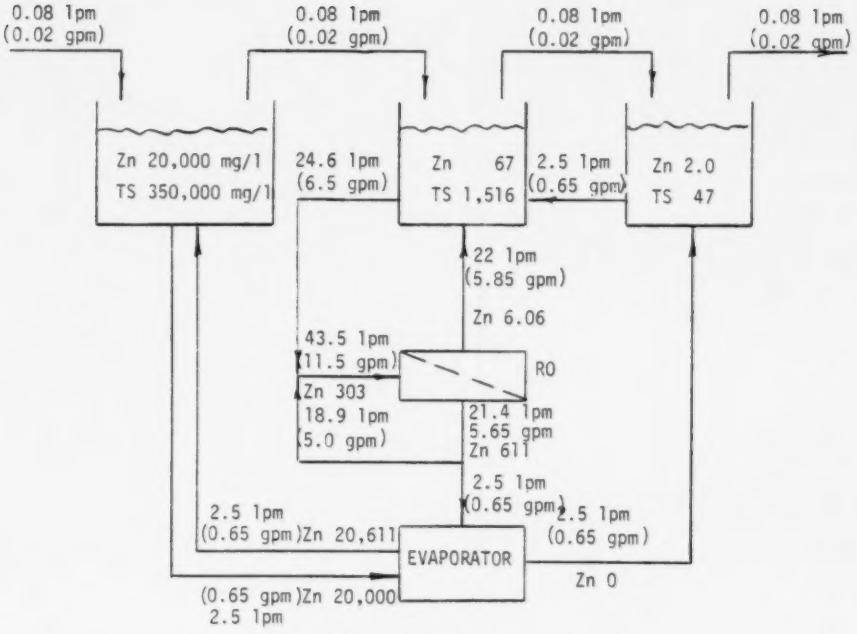


Figure 9. Dependence of rejection on feed concentration.

BEGIN FIRST LINE OF TEXT HERE
TYPING GUIDE SHEET
CENTER OF PAGE
TOP OF IMAGE AREA
DROPPED HEAD. BEGIN SECTIONS HERE



a. Present rinsing system



b. Closed-loop rinsing system

Figure 10. Flows and Concentrations for Open and Closed-loop Rinsing.

BEGIN LAST LINE OF TEXT
BOTTOM OF IMAGE AREA. OUTSIDE DIMENSION FOR TABLES AND ILLUSTRATIONS
3/8"

RO system is converted to distillate in the evaporator, and an equal flow (2.5 lpm) of plating bath is used to purge the concentrated chemicals from the evaporator and prevent precipitation. The stream returning to the bath is about 3% higher in concentration than the bath, and precipitation in the evaporator would occur only if there were some constituent in the bath (e.g., carbonates) at a concentration very close to its solubility limit. The permeate and distillate flow rates given in Figure 10b (22 and 2.5 lpm, respectively) are considerably greater than the corresponding flows in the system demonstrated (3.8-5.6 and 0.8 lpm, respectively).

One of the most important of design criteria is the flux for the RO system. The flux depends on the operating pressure, temperature, feed concentration, and the flow rate of concentrate from the module. The system would be designed to operate at 55.4 atm abs. (800 psi), maximum recommended operating pressure for the modules. The design temperature would be 25°C (77°F). At higher temperatures the flux would increase, and the rejection would remain essentially the same. Thus the performance of the RO system would improve with increasing temperature, although the membrane life may be shortened. For the design case shown in Figure 10b (i.e. 90% RO system conversion) the total solids concentration in the feed to the first RO module is calculated to be 5,500 mg/l and the concentrate withdrawn from the final module is 10,600 mg/l. Thus the average feed-side total solids concentration is about 8,000 mg/l. From Figure 8 the flux for a new module at this feed concentration would be about 34 l/m^2hr (at 47.6 atm and a recirculation flow rate of 37.9 lpm (10 gpm)). From Figure 4 the flux declines to about half of its initial value over an operating period of about 700 hours (exposure time: 3,000 hours). Thus it would seem reasonable to design the system for a minimum flux of 37.9 lpm (10 gfd). At this flux the rate of permeate production is approximately 1.9 lpm (0.5 gpm) for each module (area = 6.5 m^2(70 ft^2). Based on the recommended ratio of concentrate flow to permeate flow (10 to 1) the rate of recirculation for the RO system would be 5 gpm as shown in Figure 10b. (This recirculation rate is only half of that used during the field demonstration and could result in a slightly lower flux than measured; however the higher operating pressure -- 55.4 vs. 48.6 atm abs. (800 vs. 700 psi) -- would compensate by increasing the flux).

Rejection is also an important parameter in the design. Based on the results of Figure 9 at a total solids concentration of 8,000 mg/l, the zinc and total solids rejections would be 99% and 95%, respectively. From Figures 5 and 7, fouling of the modules can be expected to result in a decrease in zinc rejection from 99 to 98%. Thus, for the design, the minimum zinc and total solids rejections were selected as 98% and 90%, respectively.

The RO system can be designed to operate at any desired conversion within reason. At low conversion the capital and operating costs are dominated by the evaporator; at high conversion, capital and operating costs are dominated by the RO system. At some intermediate conversion, the total operating cost should pass through a minimum. Material balance calculations were performed for the system of Figure 10b using various

RO system conversions. (Results are shown only for the 90% conversion case.) Capital and operating costs were then developed for each conversion in order to determine the optimum.

Table 2 gives the capital costs for the RO unit, the evaporator unit, and the total system, for various RO system conversions. The material balance relationships and the RO module rejection were used to calculate the permeate flow and the evaporator capacity required to give a zinc concentration of 2 mg/l in the final rinse. At zero conversion in the RO system (i.e. no RO system) the entire treatment load is handled by the evaporator. For this case the evaporator must produce 7.6 lpm (2.0 gpm) in order to provide the same degree of rinsing as the original 2-stage countercurrent system. Cost details are noted at the foot of Table 2. The lowest capital cost occurs when the evaporator is used to handle the entire treatment load. For RO system conversions between 70 and 90%, the total system cost remains at about $62,000.

Total annual operating costs were also calculated for each RO system conversion. The breakdown of operating costs for a conversion of 90% is shown in Table 3. Specific costs and assumptions are noted at the foot of Table 3. The costs are based on operation 100 hours per week (Monday morning through Friday afternoon) and 50 weeks per year for a total of 5,000 operating hours per year. Similar operating costs were developed for other RO system conversions.

The electrical costs for the RO system are strongly dominated by the requirements for the high pressure pump. The use of a positive displacement pump was assumed with a combined pump/motor efficiency of 75%. In addition, since a new (or clean) module would produce twice the design flux, the RO system could be operated at half the design pressure (i.e., 28.2 atm abs. (400 psi)). As fouling occurs the operating pressure would be increased to maintain the design flux, and the system would be cleaned when the operating pressure reached 55.4 atm abs. (800 psi). Thus an average feed pressure of 41.8 atm abs. (600 psi) was used in calculating the power costs of Table 3.

When operating with the 1μ cartridge pre-filters, it was necessary to change the two parallel cartridges after about 100 hours of operation. With the 20μ filters, the pressure drop was still quite low (<3 psi) even after 300 hours of operation. With two parallel passes each having a 20μ and a 1μ cartridge in series, it is estimated that the four filters would last for about a month (400 hours). Monthly replacement was therefore assumed for the economic calculations.

With 1μ pre-filtration, cleaning was required after about 700 hours of operation. During this time three modules were in use. However for the design cases shown in Table 2, anywhere from 6 to 18 modules would be used. It is reasonable to postulate that the extent of fouling would vary inversely with the membrane surface area. Thus a cleaning frequency of once every four months (1600 hours) was assumed.

TABLE 2. CAPITAL COSTS FOR VARIOUS RO SYSTEM CONVERSIONS

RO System Conversion	Required Permeate Flow lpm (gpm)	Required Membrane Area[a] m^2 (ft^2)	Required No. of Modules[b]	Membrane Module Cost[c] $	Housing Cost[d] $	Total Cost for RO System[e] $	Required Evaporator Capacity[f] gph	Total Cost for Evaporator $	Total System Cost $
0							120	44,129	44,129
0.70	9.75 (2.57)	34.5 (371)	6	3,780	1,700	21,780	66.2	39,199	60,979
0.80	13.72 (3.62)	48.5 (522)	8	5,040	2,550	23,890	54.4	39,199	63,089
0.90	22.14 (5.85)	78.2 (842)	12	7,560	3,400	28,560	39.0	33,880	62,440
0.95	33.7 (8.91)	119.2 (1,283)	18	11,340	5,100	34,040	28.1	33,880	67,920

(a) Design flux = 17 L/m^2-hr (10 gfd).

(b) Based on area of 6.5 m^2 (70 ft^2) per module.

(c) Based on $630 per module (Abcor, Inc.).

(d) Based on $850 per 3-module housing (The Permutit Co., Inc.).

(e) Based on system cost of $15,000 (Osmonics, Inc.) for system W/O modules, housings, and high-pressure pump. Pump/motor cost = $1,300 for <15 Lpm permeate; $2,600 for >15 Lpm permeate (Wanner Engineering, Inc.).

(f) Double effect evaporator with cooling tower package. Based on rated capacities of 757 Lph (200 gph) ($44,139), 379 Lph (100 gph)($39,199), and 189 Lph (50 gph)($33,880).

TABLE 3. TOTAL AND NET OPERATING COSTS FOR 90% RO SYSTEM CONVERSION (5,000 OPERATING HOURS PER YEAR)

Item	RO System $	Evaporator System $	Total System $
Capital Amortization [a]	2,100	3,388	5,488
Direct Operating Costs			
Steam @ $3.50/1000 lbs [b]	-	3,760	3,760
Electrical @ $0.0474/kw-hr [c]	1,008	1,100	2,108
Operation & Maintenance @ $10/hr including fringe benefits [d]	960	960	1,920
Cartridge Filters & Cleaning Chemicals [e]	240	-	240
Membrane Replacement (2-year life)	3,780	-	3,780
Total Direct Operating Cost	5,998	5,820	11,808
Total Annual Operating Cost	8,088	9,208	17,296
Credit for Bath Recovery @ $0.47/gal [f]			(2,820)
Credit for Water Recovery @ $0.45/100 ft^3			(361)
Credit for Cyanide Destruction @ $4.05/lb CN [g]			(6,440)
Credit for Solid Waste Disposal . $29/drum and 25% solids in sludge [h]			(354)
Total Annual Credits			(9,975)
Net Annual Operating Cost			$7,321

(a) Straight-line depreciation over 10 years with zero salvage value.

(b) Based on actual cost for No. 4 fuel oil of $0.393/gal, heating value of 140,000 Btu/gal, and a boiler efficiency of 80%.

(c) Based on actual cost for October 1978.

(d) Based on actual maintenance labor rates.

(e) Four cartridge filters changed once per month at average cost of $4.68 each. Cleaning three times per year with 6 lbs citric acid per cleaning at $0.82 per pound.

(f) Chemical costs: zinc $0.445/lb; sodium cyanide $0.51/lb; caustic $0.185/lb; brightener $6.792/gal.

(g) Based on 8 lbs NaOCl per lb of CN with no addition of caustic (already present in rinsewater). Cost for 15% NaOCl solution = $0.665/gal.

(h) Based on sludge centrifuged to 25% solids and hauled in lots of 80 55-gal drums.

Membrane life is a very important parameter in the economics. As mentioned above, chemical attack of the membrane or other module components would be expected to depend on total time of exposure to the rinsewaters rather than operating time per se. The two field demonstration modules were exposed to the rinsewaters for a total of 4,200 hours. This is close to the 5,000 hours operating time per year taken as the basis for calculating the operating costs. If the system were designed to purge the modules with distillate on shut-down, the effective exposure time would also be about 5,000 hours. Thus a membrane life on one year would be virtually assured. However, it is likely that the membrane life would be considerably longer than one year. During the field demonstration the modules produced only a fraction of the permeate flow that would be produced by a system designed to meet the rinsing constraints of Figure 10b. Thus the feed concentration to the field demonstration modules was considerably greater than would be seen by a larger capacity system. Furthermore, during periods when the demonstration system was not operating, the concentration across the membranes equilibrated resulting in very high concentrations on the permeate side. This situation would accelerate attack of the membrane backing material and other permeate-side components. Purging the system on shut-down would largely eliminate this source of module deterioration. Considering these factors, it is not unreasonable to project a two-year membrane life as assumed in Table 3.

Operating and maintenance labor for the RO system was assumed to require 1 day per month. This would include system start-up on Mondays, shut-down on Fridays, cartridge filter replacement, membrane cleaning, and other maintenance as required.

Steam consumption for a double effect evaporator operating as shown in Figure 10b with temperatures of 54°C (130°F) and 43°C (110°F) in the first and second effects, respectively was calculated to be 0.66 lbs steam/lb distillate. Electrical requirements for operation of the cooling tower, evacuation of the evaporator, pumping the various output streams, and miscellaneous usage were obtained from the manufacturer (4.64 kw, 8.64 kw, and 16.64 kw for double effect capacities of 189 lpm, 379 lpm, and 757 lpm, respectively). As for RO, operation and maintnenace for the evaporator was assumed to require one man-day per month.

As shown in Table 3, the direct operating costs (excluding amortization) are about $6,000 per year each for the RO system and the evaporator for a total of about $12,000. The total annual operating cost including amortization is $17,300. Table 3 also gives the breakdown of credits resulting from recovery of the rinsewater. The largest credit results from a reduction in the amount of cyanide to be destroyed. The credit for sludge disposal was calculated on the basis of the weight of zinc hydroxide produced by precipitation of the dragout and is probably significantly less than for plating shops which do not have solids dewatering equipment. The total annual credit for rinsewater recovery is about $10,000, which leaves a net operating expense of $7,300 per year for the recovery system.

Operating costs were also determined for other RO system conversions and are shown in Figure 11. For conversions ranging from 70% to 95% the RO operating costs increase and the evaporator operating costs decrease. The total operating cost passes through a minimum at an RO system conversion of 90%. Using an evaporator alone for rinsewater recovery (zero percent RO system conversion) the total annual operating cost including amortization is about $20,900. For the optimum combination of RO and evaporation, the total annual operating cost is approximately $17,300 which represents an annual savings of $3,600.

In considering the impact of these numbers it should be emphasized that the recovery system was designed to meet a given rinsing criteria for a two-stage countercurrent rinsing system. Recovery system costs could be reduced significantly by using more rinsing stages. For manual operations, over-the-tank spray rinses could be used. In addition, consideration should be given to working with higher rinse concentrations where rinsing is not critical. The costs for recovery provide a substantial incentive for reducing the rinsewater flow by simple, inexpensive techniques.

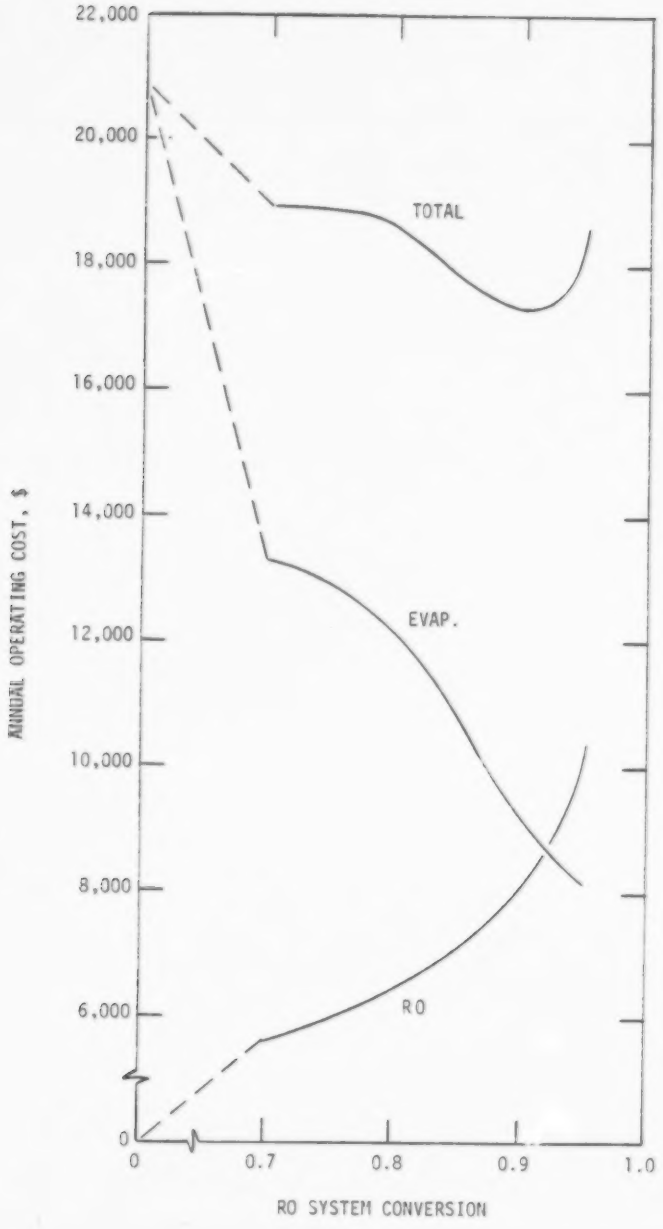


Figure 11. Annual Operating Costs for Various RO Conversions.

REFERENCES

1. Skovronek, H. S., and M. K. Stinson. Advanced Treatment Approaches for Metal Finishing Wastewaters (Part II). Plating and Surface Finishing, 64 (11): 24-31, 1977.

2. Anonomous. Recovery Pays! Plating and Surface Finishing, 66 (2): 45-48, 1979.

3. Hall, E. P., D. J. Lizdas, and E. E. Auerbach. Plating and Surface Finishing, 66 (2): 49-53, 1979.

4. Donnelly, R. G., R. L. Goldsmith, K. J. McNulty, and M. Tan. Reverse Osmosis Treatment of Electroplating Wastes. Plating, 61 (5): 432-422, 1974.

5. McNulty, K. J., R. L. Goldsmith, A. Gollan, S. Hossain, and D. Grant. Reverse Osmosis Field Test: Treatment of Copper Cyanide Rinse Waters. EPA-600/2-77-170, U.S. Environmental Protection Agency, Cincinnati, Ohio, 1977. 89 pp.

6. McNulty, K. J., P. R. Hoover, and R. L. Goldsmith, Evaluation of Advanced Reverse Osmosis Membranes for the Treatment of Electroplating Wastes. In: First Annual Conference on Advanced Pollution Control for the Metal Finishing Industry. EPA-600/8-78-010, U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. pp 66-75.

7. Personal Communications, Dr. Robert L. Riley, Fluid Systems Division of UOP, San Diego, CA.